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Technical and economic assessment of ammonia-based post-combustion CO_2 capture

Peter Versteeg[a], Edward S. Rubin[a]

[a]Department of Engineering and Public Policy, Carnegie Mellon University, 5000 Forbes Ave. Pittsburgh, 15213, USA

Abstract

The performance and cost of two ammonia-based post-combustion CO_2 capture systems operating at a new supercritical coal-fired power plant were modeled and compared to an amine-based CO_2 capture system operating at a similar plant. This assessment showed that for a fixed coal input, the plant derating of a CO_2 capture system operating with high ammonia concentrations (HighNH_3) was found to be 2 percentage points lower than a plant with the amine-based system. The plant derating of a CO_2 capture system operating with low ammonia concentrations (LowNH_3) was substantially higher. Preliminary estimates of the revenue requirement of the plants with HighNH_3 and LowNH_3 systems are $U.S. 117/MWh and $U.S. 148/MWh respectively, compared to $U.S. 119/MWh for a plant with an amine-based system. The results from this performance assessment and preliminary cost analysis suggest that the LowNH_3 system will not be competitive and that the HighNH_3 system may have a slight energy and cost advantage over the amine system. Furthermore, a preliminary uncertainty analysis explores the critical factors that may affect the performance and cost estimates of these systems, including the potential for slow reaction kinetics to increase absorber costs, and these results are presented.

Keywords: Post-combustion; chilled-ammonia; carbon capture and sequestration

1. Introduction

Policy makers face central questions about costs and effectiveness in assessing CO_2 emission mitigation options [1]. One such option is post-combustion CO_2 capture, which targets CO_2 emissions released from the burning of fossil fuels. Amine scrubbing is the leading post-combustion CO_2 capture technology, and is well understood and ready for large scale use [2]. Post-combustion CO_2 capture based on ammonia is less understood but is attractive because ammonia is inexpensive, the CO_2 can be regenerated at high pressure, and the steam requirements for the regeneration process may be lower than for amine-based technologies. This study presents a performance assessment and preliminary cost analysis of a power plant integrated with ammonia-based CO_2 capture, and compares the results to a plant with an amine-based system. Using tools that have already been developed, this paper is intended to be a starting point for estimating costs for this process, and will help policy makers be more informed about the costs of CO_2 emission mitigation options.

For comparing performance and costs between post-combustion CO_2 capture with ammonia and amine technologies, this paper uses the plant derating of CO_2 capture on the power plant and the levelized revenue required as two key parameters, as calculated in equation 1 and equation 2. The plant derating for CO_2 capture is expressed as the percentage reduction in net plant output for a constant energy input and is occasionally reported as an "energy penalty".

$$\text{Plant Derating (\%)} = \frac{\text{Plant Efficiency without Capture} - \text{Plant Efficiency with Capture}}{\text{Plant Efficiency without Capture}} \quad (1)$$

$$\text{Revenue Required (\$/MWh)} = \frac{\text{Total Plant Costs} * \text{Fixed Charge Factor} + \text{O\&M Costs}}{8760 * \text{Capacity Factor} * \text{MWh Produced}} \quad (2)$$

* Corresponding author. Tel.: +1-412-567-7785
E-mail address: plv@andrew.cmu.edu.

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Plant derating estimates of amine and ammonia-based CO_2 capture systems from previous studies are shown in Table 1. Early plant derating estimates of ammonia-based CCS are lower because initial studies assumed that an ammonia-based capture system could be designed around the low energy reaction between ammonium carbonate ($(NH_4)_2CO_3$) and ammonium bicarbonate (NH_4HCO_3). More recent studies have higher plant derating estimates because the formation of other chemical species was also found to be important [3].

Table 1: Comparison of plant derating estimates for amine and ammonia-based CO_2 capture systems.

Study Type	Plant Derating	Authors, Publication, Affiliation & Year	Notes	Ref
Amine				
Performance and Cost Estimate	28%	Buchanan et al., EPRI 2000	Calculated from data	[4]
Performance and Cost Estimate	30%	Woods et al., NETL 2007	Calculated from data	[5]
Vendor Estimate	23%	Kishimoto et al., MHI 2009	Estimated from data	[6]
Expert Elicitation	25-28%	Chung et al., 2009	Power plant retrofit in 2030	[7]
Ammonia				
Performance and Cost Estimate	17%	Ciferno et al., NETL 2005	Calculated in study	[8]
Performance and Cost Estimate	10-14%	Gal, EPRI 2006	Calculated from data	[9]
Vendor Estimate	9%	Peltier, Powermag 2008	Calculated from data from Alstom	[10]
Performance Estimate	13%	Valenti et al. 2008	Calculated from data	[11]
Performance Estimate	28%	Mathias et al., Fluor 2008	Estimated from data	[3]
Vendor Estimate	17%	McLarnon et al., Powerspan 2009	Calculated from data	[12]
Vendor Estimate	20%	Hilton et al., Alstom 2009	Estimated by Alstom	[13]
Expert Elicitation	17-20%	Chung et al. 2009	Power Plant Retrofit in 2030	[7]

2. Methodology

The characteristics of the power plant used in this study are derived from the 2007 DOE/NETL Bituminous Baseline report (Case 12) [5]. This plant is a pulverized coal, supercritical Rankine cycle plant that burns Illinois No. 6 coal and is located in the Midwestern USA. The plant is fitted with a Fluor Econamine FG Plus process, and in this study the flue gas into this system is used as the feed for the ammonia-based systems.

To simulate CO_2 capture using ammonia, an equilibrium electrolyte model developed for this purpose by Aspen Technologies is used [14]. Aqueous ammonium bicarbonate (NH_4HCO_3), ammonium carbonate ($(NH_4)_2CO_3$), and ammonium carbamate (NH_2COONH_4) as well as the salt precipitate of ammonium carbonate can occur, depending on process conditions. Salt precipitation lowers the concentration of aqueous ammonium bicarbonate allowing additional CO_2 to be absorbed. Multiple ammonia-based CO_2 capture systems can be designed by taking advantage of different aspects of the chemistry.

The basic process steps and conditions for ammonia-based CCS are taken from a patent application by Gal [15], a presentation by Hilton et al. [13], and work by Mathias et al. [3], and the major features are modelled in Aspen Plus® (V7.1) as shown in Figure 1. Power plant flue gasses are initially cooled using circulating water and a direct contact cooler, and most of the water in the gases are condensed out. The flue gases are further cooled in a cross flow heat exchanger using chilled water from a vapour compressor. The chilled flue gasses feed into a CO_2 absorption column at 10°C and 1 atm, where the gases are contacted with a lean solvent mixture. The lean solvent contains ammonia, carbon dioxide, and water, and has a lean loading NH_3/CO_2 ratio of 2.85. As shown in Figure 2a, higher ratios result in increased ammonia slip over the absorber and therefore increased flue gas cleaning demands, while lower ratios require additional solvent flow or higher ammonia concentrations to capture 90% CO_2. The ratio of 2.85 was chosen as a compromise between these two issues. The CO_2 rich stream leaves the bottom of the absorber and is compressed to 3.0 MPa by a high pressure pump. The rich solution then flows through a cross flow heat exchanger where it is heated by the hot lean solution coming off the reboiler, and if any solids remain in the stream a heater is used to dissolve them. Steam is used in the CO_2 stripper to regenerate the CO_2 at 2.8 MPa, and the regenerated solvent is then returned to the absorber. The absorber gasses are cleaned of ammonia in a water wash and are then heated in a second direct contact cooler, before being released through the stack. Finally, distillate from the second stripper containing ammonia, carbon dioxide, and water is fed back to the CO_2 absorber. Several components were modelled separately in Excel®, including the CO_2 compressor equipment and the water chillers that supply cooling loads.

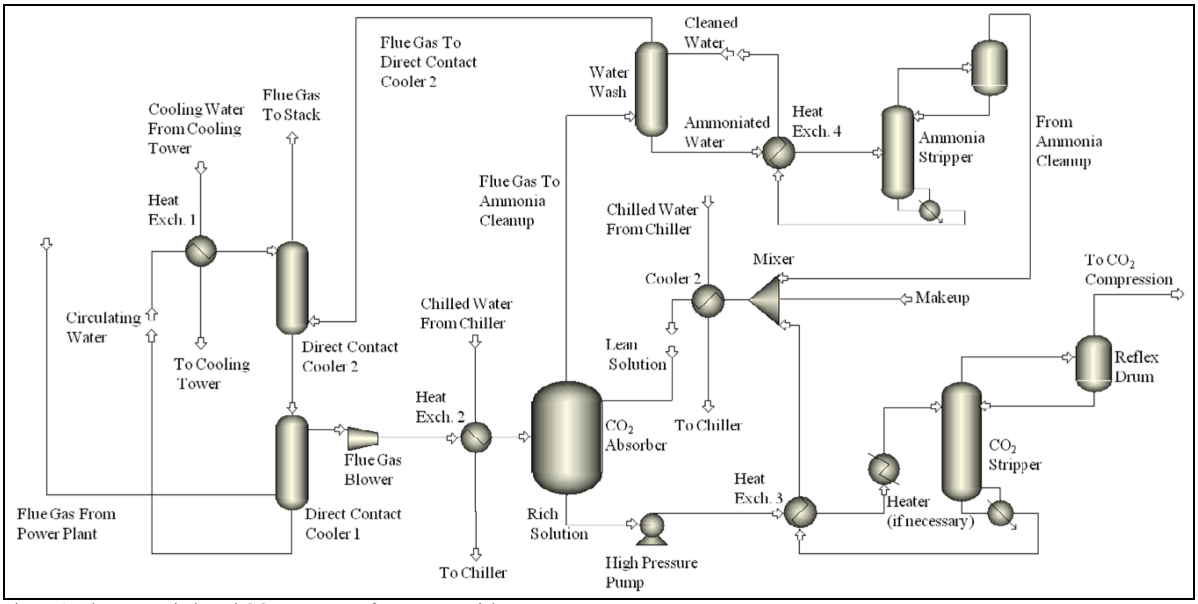

Figure 1: The ammonia-based CO_2 capture performance model.

The ammonia concentrations and solvent flow rates in the CO_2 capture process affect performance and equipment cost. The impact of varied ammonia concentrations over a consistent design is absent in the literature and so we present a sensitivity analysis of the effect of ammonia concentration and solvent flow rate on CO_2 absorption. The results for percent CO_2 capture in the absorber, absorber ammonia slip, and solids content out of the bottom of the absorber are shown in Figure 2b, 2c, and 2d respectively. Increases in CO_2 capture results in higher ammonia slip, while the use of low flow rates or high ammonia concentrations cause ammonia salts to precipitate in large quantities.

From the results of the sensitivity analysis two CO_2 capture system designs were considered, a low concentration ammonia system operating without solids (LowNH$_3$) and a high concentration system operating with a rich solvent absorber outlet stream of 60 wt% solids (HighNH$_3$). Additional lean solvent ammonia would result in salt precipitation for LowNH$_3$, while HighNH$_3$ has a rich solvent solids content equivalent to the highest solids wt % for ammonia-based CCS found in the literature [3]. The designs for both systems include flue gas cooling, absorber cooling, and ammonia cleanup of the flue gas. For both LowNH$_3$ and HighNH$_3$, the key design variables, process conditions, and predicted flows are shown in Table 2. The minimum temperature approach for the heat exchangers in both systems is 5.6°C, with the exception of Heat Exch. 3 for LowNH$_3$, which has a temperature approach of 20°C to control costs for this piece of equipment.

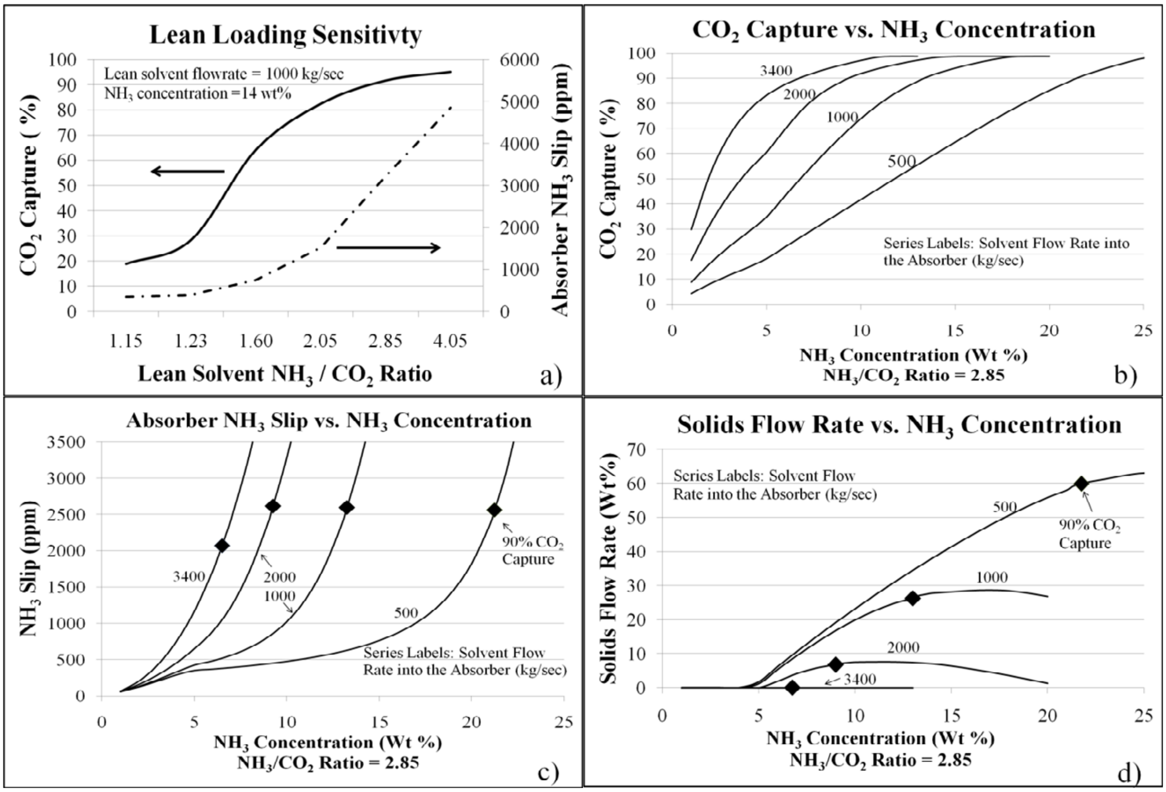

Figure 2: a) Lean loading vs. absorber CO_2 capture and absorber NH_3 slip, b) lean solvent NH_3 concentration vs. CO_2 capture, c) lean solvent NH_3 concentration vs. absorber NH_3 slip, and d) lean solvent NH_3 concentration vs. solids flow rate. Diamonds represent 90% CO_2 capture.

Table 2: Key process conditions and flows for the ammonia-based CO_2 capture system designs.

Parameter	$LowNH_3$	$HighNH_3$
Flue Gas Flow Rate into the System (kg/sec)	860	860
Flue Gas CO_2 Mole Fraction	13.3	13.3
Flue Gas Circulating Water Flow Rate (kg/sec)	1452	1452
Flue Gas Water Wash Cleaning Water Flow Rate (kg/sec)	32	36
Flue Gas HeatX 2 Chilling Load (10^3 tons cooling/ hr @ 3°C)	10	10
CO_2 Absorber Solvent Flow Rate (kg/sec)	3,400	500
CO_2 Absorber Lean Solvent NH_3 Concentration (wt%)	6.75%	21.5%
CO_2 Absorber Chilling Load (10^3 tons cooling/ hr @ 3°C)	70	107
CO_2 Absorber CO_2 Removal Efficiency (%)	90%	90%
CO_2 Absorber Ammonia Slip (ppm)	2112	2488
CO_2 Absorber Rich Stream Solids Content (wt%)	No Solids Occur	60%
Lean Solvent Chilling Load (10^3 tons cooling/ hr @ 3°C)	88	5
Overall Ammonia Slip (ppm)	<1	<1
Overall CO_2 Capture (%)	90%	90%
Overall CO_2 Product Purity (vol%)	99.8%	99.8%

3. Power Plant Performance

The power usage of both $LowNH_3$ and $HighNH_3$, including the electrical equivalent in steam drawn off after the intermediate turbine, is shown in Table 3 along with the resulting performance characteristics of the power plant. The performance estimates are based primarily on performance data from Aspen Plus® as well as data scaled from the Integrated Environmental Control Model (IECM V6.2), a program which can provide performance and cost estimates for power plants [16]. The plant derating of the $HighNH_3$ was found to be 2 percentage points lower than the plant derating of the amine-based system. The plant derating of $LowNH_3$ however, was found to be 9 points higher. This suggests that the $LowNH_3$ system will not be competitive. $HighNH_3$ performs relatively well due to the energy benefits associated with higher CO_2 loading and reduced heating, cooling, and transportation energy requirements.

Table 3: Power plant performance estimates. All values are in MWe equivalent

	No CO_2 Capture [5]	Amine System [5]	$LowNH_3$	$HighNH_3$	Notes and Primary Data Sources for Calculation
Potential Power Available	580.2	827.6	827.6	827.6	Based on coal flow rate
Auxiliary Steam Load					
Heater				14.8	Aspen Plus
CO_2 Stripper		164.2	163.0	87.0	Aspen Plus
NH_3 Stripper			3.5	3.6	Aspen Plus
Steam Turbine Power	580.2	663.4	661.1	722.2	Based on aux. steam load
Auxiliary Electrical Load					
Flue Gas Blower			18.9	18.9	ΔP=3 psi, scaled IECM data
Gas Cooling Water Pumps			3.2	3.2	Aspen Plus, scaled IECM data
Chiller for Heat Exch 2			5.7	5.7	Aspen Plus, [17], [18]
Chiller for Absorber Cooling			38.4	58.9	Aspen Plus, [17], [18]
Chiller for Solvent Cooling			48.5	3.0	Aspen Plus, [17], [18]
Absorber Cooling Pumps			3.5	5.4	Aspen Plus, scaled IECM data
Solvent Circulation Pumps			1.5	0.2	Aspen Plus, scaled IECM data
Econamine FG Plus System		23.2			
CO_2 Compression		46.9	17.0	17.0	Aspen Plus, scaled IECM data, [5]
Balance of Plant	30.1	49.2	49.0	49.0	Scaled IECM data
Plant Net Power	550.1	546.0	475.2	560.8	
Plant Efficiency (% HHV)	39.1%	27.2%	23.7%	28.0%	
Plant Derating of CO_2 Capture (%)		30.4%	39.4%	28.5%	Equation 1

4. Power Plant Costs

Preliminary cost results for the two ammonia-based CO_2 capture system designs in 2007 constant dollars are shown in Table 4. The cost estimates are based on data from the IECM, sources from the literature, and Aspen Icarus®. The costs for a number of components are scaled according to the methodology described in [19]. All the plants in Table 6 are assumed to have a levelized capacity factor of 75% over their lifetimes. The capital costs for both $HighNH_3$ and $LowNH_3$ are higher than for the amine system. Revenue required estimates for the plants with $HighNH_3$ and $LowNH_3$ systems are $U.S.117/MWh and $U.S.148/MWh respectively and compare to $U.S.119/MWh for the plant with amine-based CO_2 capture. The plant with $HighNH_3$ has a higher efficiency, which leads to a slight cost advantage compared to the plant with amine-based CO_2 capture. In comparison with $LowNH_3$, $HighNH_3$ again benefits from smaller equipment sizes associated with higher loading and reduced heating, cooling, and transportation requirements.

The absorber in $HighNH_3$ is required to handle significant amounts of solids and was therefore considered a spray tower with capital costs similar to that of a wet flue gas desulfurization system. This cost estimate may be optimistic because modeling is based on equilibrium assumptions which may not apply in all cases. Given that a very large absorber would be required for a close approach to equilibrium, and that previous investigations on the kinetics of ammonia based CO_2 capture have shown that absorption may be slower than for monoethanolamine (MEA) based CO_2 capture [20], the absorber for this system may be significantly more expensive than the estimate provided here. This issue is considered further in the uncertainty analysis in Section 5.

Table 4: Power plant cost estimates with ammonia-based CO_2 capture, values are in 2007 $Millions

	No CO_2 Capture[1]	Amine System[1]	LowNH_3	HighNH_3	Notes and Primary Data Sources for Calculation
CO_2 Capture Process Area Costs					
DCC #1			30.9	30.9	Aspen Plus, scaled IECM data, [19]
DCC #2			23.3	23.3	Aspen Plus, scaled IECM data, [19]
Flue Gas Blower			6.4	6.4	Aspen Plus, scaled IECM data, [19]
Heat Exch. 1			6.7	6.7	Aspen Icarus
Heat Exch. 2			2.9	2.9	Aspen Icarus
Heat Exch. 1 Pumps			1.4	1.4	Aspen Plus, scaled IECM data, [19]
Heat Exch. 2 Pumps			0.5	0.5	Aspen Plus, scaled IECM data, [19]
Cooling Water Circ Pumps			0.7	0.7	Aspen Plus, scaled IECM data, [19]
Chiller System			74.3	54.3	Aspen Plus, [18]
Absorber			74.4	105.1	Aspen Plus, scaled IECM data
Absorber Pumps			1.9	2.5	Aspen Plus, scaled IECM data, [19]
Heat Exch. 3			74.7	19.2	Aspen Icarus
Solvent Circulation Pumps			16.5	5.2	Aspen Plus, scaled IECM data, [19]
Solvent Heater			0.0	2.5	Aspen Icarus
Solvent Cooler			37.5	2.3	Aspen Icarus
CO_2 Stripper			66.5	21.0	Aspen Plus, scaled IECM data, [19]
CO_2 Stripper Reboiler			33.2	7.2	Aspen Plus, scaled IECM data, [19]
Water Wash			2.2	2.2	Aspen Icarus
Heat Exch. 4			0.1	0.1	Aspen Icarus
NH_3 Stripper			1.5	1.5	Aspen Icarus
NH_3 Cleanup Pumps			1.0	1.1	Aspen Plus, scaled IECM data, [19]
Steam Extractor			3.3	3.3	Scaled IECM data
Sorbent Processing			1.1	1.1	Scaled IECM data
Drying and Compress Unit			18.3	18.3	Aspen Plus, scaled IECM data
General Facilities Capital			7.5	5.0	1.57 % PFC, [5]
Eng. & Home Office Fees			45.0	30.1	9.37 % PFC, [5]
Project Contingency Cost			78.5	52.6	16.38 % PFC, [5]
Process Contingency Cost			22.4	15.0	4.67 % PFC, [5]
CO_2 System (TCR)		393.9	633.2	424.0	Based on Area Costs
Base Plant (TCR)[2]	670.8	881.3	865.4	884.1	Scaled IECM data
Cooling Tower (TCR)	35.8	62.7	62.7	62.7	Scaled IECM data
NO_x Control (TCR)	25.0	33.7	33.7	33.7	Scaled IECM data
TSP Control (TCR)	37.4	49.8	49.8	49.8	Scaled IECM data
SO_2 Control (TCR)	112.1	138.7	138.7	138.7	Scaled IECM data
CO_2 System and TS&M O&M/Year		24.1	21.6	21.6	Scaled IECM data
Balance of Plant O&M/Year	103.1	128.9	128.9	128.9	Scaled IECM data
Plant Total Capital Requirement	881.1	1560.0	1783.5	1593.0	Based on TCR Costs
Total O&M Costs/Year	103.1	153.0	150.6	150.6	Total O&M
Capital Required ($/kW-net)	1601.0	2857.0	3753.3	2840.5	Based on Performance
Revenue Required ($/MWh)	60.4	118.7	148.2	116.5	Equation 2

[1]The plants with amine-based CO_2 capture are based on Case 11 and Case 12 in [5], and were modelled in the IECM with a 75% capacity factor. [2]The base plant cost is reduced for the low concentration ammonia-based CO_2 capture system design because a smaller steam turbine is required.

5. Uncertainty Analysis on Key CO2 Capture System Performance Parameters

The performance and cost estimates have large uncertainties due to the limited availability of pilot plant data for model validation, the proprietary nature of industrial process designs now under development, and the limited commercial experience with building and operating commercial power plants with CCS. A preliminary uncertainty analysis using Monte Carlo simulation was used to illustrate the effect of uncertainty of key CO_2 capture system variables on the overall performance of the power plant. The uncertainty in each of the system variables was estimated from literature data where available, and probability distributions were assigned to the variables based on these estimates, as shown in Table 5. The probability distribution for the CO_2 absorber costs is weighted to reflect that slower rates of reaction may negatively affect these costs. The stochastic results are shown in Figure 6.

Table 5: Key uncertainties associated with the CO_2 capture systems.

Parameter	Units	Nominal	Distribution Function
Solvent Chilling Loads		Table 3	Uniform(-25%,+25%)
Auxiliary Steam Loads		Table 3	Uniform(-25%,+25%)
ΔP Across Capture System	psi	3	Uniform(2, 4)
Chiller Electrical Use, 3°C Water Product	kW/ton	0.55	Triangular(0.50, 0.55, 0.60)
CO_2 Compression, from 27.5 bar to 152.7 bar	kWh/kg CO_2	0.03	Triangular(0.028, 0.03, 0.032)
Cooling Equipment Costs	$/ton cooling	441	Uniform(-30%, +30%)
IECM Based Equipment Costs	$	Table 4	Uniform(-30%, +30%)
Aspen Icarus® Based Equipment Costs	$	Table 4	Uniform(-40%, +40%)
CO_2 Absorber Costs	$	Table 4	Uniform(-30%, +250%)
Power Plant Fixed Charge Factor	--	0.175	Uniform(0.16,0.19)
Power Plant Levelized Capacity Factor	--	0.75	Weibull(8.5, 0.81)

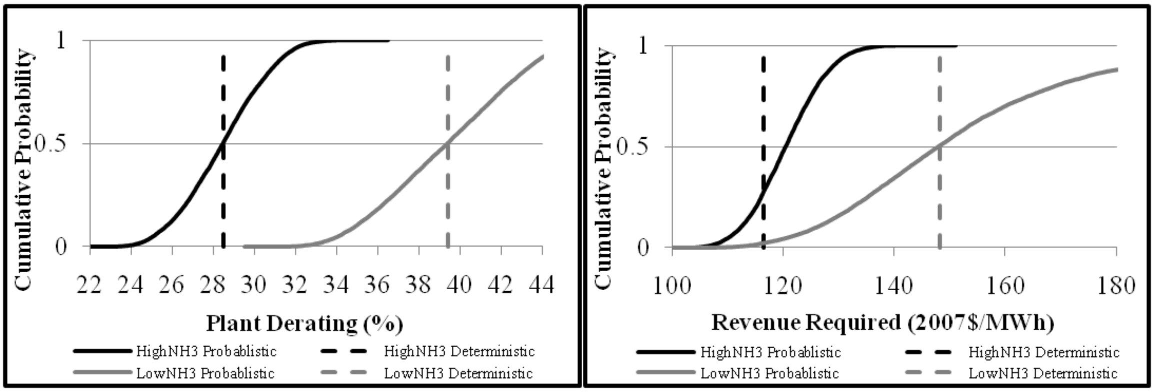

Figure 6: Cumulative probability distribution of plant performance (as reflected by the plant derating on the left) and total plant costs (levelized revenue requirement, on the right) of HighNH$_3$ and LowNH$_3$. The deterministic results are shown as a vertical line.

6. Discussion and Conclusion

The plant derating and equipment costs for low temperature CO_2 absorption are substantial for both LowNH$_3$ and HighNH$_3$. The cost estimates are higher for LowNH$_3$ due to larger equipment sizes, the energy requirements associated with high flowrates, and the extensive cooling loads required. The HighNH$_3$ system may have a slight energy and cost advantage over amine systems. A preliminary uncertainty analysis explored the critical factors that may affect the performance and cost estimates of these systems, and these results are presented.

The intent of this study was to provide reasonable preliminary performance and cost estimates of ammonia-based CO_2 capture system designs and relative comparisons with amine-based CO_2 capture systems. In the future the ammonia-based CO_2 capture performance and cost models could benefit from improved thermodynamic models, more detailed simulations of individual pieces of equipment including in particular the absorber, rate-based as opposed to equilibrium modelling, and cost estimates by vendors.

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